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PATENT SPECIFICATION

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COMPLETE SPECIFICATION

Petroleum Oil Refining Process

We, STANDARD OIL DEVELOPMENT COMPANY, a Corporation duly organized and existing under the laws of the State of Delaware, United States of America, having an office at Elizabeth, New Jersey, United States of America, do hereby declare the invention, for which we pray that a patent may be granted to us, and the method by which it is to be performed, to be particularly described in and by the following statement:—

This invention is concerned with a combination process for the conversion of crude petroleum into more valuable mineral oil products, particularly with obtaining maximum yields of low boiling products of the motor fuel and heating oil boiling ranges and minimum yields of heavy products, such as fuel oils and tar with a minimum of treating steps and a substantially reduced investment and operating cost.

In the combination crude oil distillation and conversion processes previously proposed the recovery of maximum yields of motor fuel and heating oil products has been accomplished usually by subjecting light and heavy naphtha fractions from the crude still to further fractionation and, if desired, to suitable thermal or catalytic refining treatments, such as reformation, isomerization, hydroforming, alkylation, or, by thermally or catalytically cracking a gas oil fraction from the crude still to recover further low boiling product by subsequent fractionation of the cracked products and subjecting the reduced crude to a further distillation at reduced pressures to produce tar and additional low boiling products, principally gas oil to be processed with the gas oil fraction from the crude still as mentioned above. These processes require a plurality of product fractionators yielding several streams of products of desirable boiling ranges. For economic heat recovery, large numbers of heat exchange apparatus are required both

within each unit and in combination between units. Vast tank facilities must be provided to permit storage of the various products prior to blending in desired proportions. The vacuum distillation equipment used for working up the reduced crude is expensive with respect to investment, operation and maintenance. As a result of these complications, conventional type combination processes must be operated on a relatively large scale to be economical. Normally, refining capacities in excess of, say, about 20,000 barrels (oil)/day of crude are required to make operations of this type pay while smaller refineries must be designed on the basis of an often undesirably high output of heavy fuel oil and other products of a relatively low commercial value.

In accordance with this invention, there is provided a combination process for the refining of petroleum oils which comprises distilling a crude petroleum oil to yield a residual fraction and a plurality of lighter fractions, subjecting at least one of said lighter fractions to a conversion treatment, passing said residual fraction to a fractionating tower and feeding a plurality of hydrocarbon streams into said fractionating tower below the point of entry thereto of said residual fraction whereby the latter is stripped by the vapors of said hydrocarbon streams, and recovering the desired product fractions from said fractionating tower, wherein said hydrocarbon streams fed to the fractionating tower comprise at least one total effluent from a conversion treatment of said lighter fractions and, if available, at least one of said lighter fractions as such.

Although, in general, the combination process of the invention is so practised that the reduced crude residual fraction from the crude oil distillation stage is passed as such to the fractionating tower the invention also include modifications

where this is not the case. In one such modification the reduced crude residual fraction is subjected to a high temperature coking or viscosity reducing treatment from which the effluent, including if desired the coke products, is then passed to the fractionating tower; and in a second modification the reduced crude residual fraction is subjected to a stripping operation to remove lighter fractions therefrom, the stripped reduced crude being then passed to the fractionating tower.

In practising one embodiment of the invention, the crude oil is subjected to distillation in a conventional crude distillation unit to produce an over-head stream of light virgin naphtha, a separate heavy naphtha stream, a still heavier stream of the kerosene and Diesel oil range and reduced crude bottoms. The reduced crude bottoms may still contain the gas oil fraction intended in part at least for feed stock to a gas oil cracking step. The kerosene fraction may be recovered as product directly from the crude still because further conversion will not substantially enhance its value as kerosene or Diesel oil. All other fractions are supplied to a single product fractionator substantially as follows.

The reduced crude obtained as above described is passed directly to an upper portion of the lower contacting section of a substantially conventional fractionating column. The heavy naphtha stream is subjected to a high temperature thermal or catalytic reformation or other conversion conducive to an improvement of its motor fuel qualities, and the total vaporized effluent from this conversion stage is fed to the fractionating column at a point below the feed point of the reduced crude, substantially at the temperature of the conversion stage. The light virgin naphtha fraction is likewise fed to the fractionating column at a point below the reduced crude feed point and can be injected in either liquid or vapor state after suitable preheating if desired. Various final product streams are recovered from the product fractionator for example a fuel gas overhead, a low boiling fraction of the motor fuel boiling range, a heating oil fraction, a gas oil fraction, and a heavy bottoms fraction of the fuel oil range. In addition, there may be produced a gas oil fraction which contains the virgin gas oil fraction desired for feed to further conversion equipment together and in admixture with a recycle or converted fraction in approximate optimum ratio for ultimate conversion operation. Most of the gas oil fraction (which may if desired be mixed with a

virgin gas oil fraction from the crude oil distillation stage) is passed to a preferably catalytic cracking stage to be converted therein into additional amounts of motor fuel, Diesel oil, gas oil range cycle stock and heavy bottoms. The total effluent of this cracking stage is likewise passed without substantial heat loss to the product fractionator at a point below the reduced crude feed, but preferably above the feed points of the reformate from the treatment of the heavy naphtha and of the light virgin naphtha. In this manner, the reduced crude is subjected to successive countercurrent heating, vaporization and stripping action, first with vapors from the gas oil cracking process which are heavier and may be at a lower temperature than the naphtha vapors used and thereafter with the low molecular weight, and, preferably, higher temperature naphtha vapors.

When operating substantially in the manner described above, extremely large volumes of process vapors are available and utilized for reduced crude stripping with the effect that the volume of heavy fuel finally produced may be kept at a minimum at least as low as and, if desired, even lower than may be accomplished in conventional operation involving vacuum distillation of reduced crude. The total product naphtha is prepared as overhead product which has been maintained at high temperatures for a minimum time and is, therefore, of superior quality with respect to engine cleanliness. A single heavy fuel fraction is obtained in which all heavy constituents formed in the various stages are combined and which may be subjected to a single final treatment to recover a final low sediment fuel oil of best quality. These advantages are in addition to the obvious savings of heat exchange equipment resulting from the elimination of various intermediate heating and cooling operations and to those of intermediate tankage requirements resulting from the use of a single product fractionation stage.

The number and type of intermediate conversion stages employed in the embodiment described above will depend entirely on the character of the crude oil to be processed and the types, quality and relative proportions of the desired products. Other conversion stages may be added to or substituted for those mentioned. For example, in accordance with a modification already mentioned, a coking or viscosity-reducing treatment may be applied to the reduced crude prior to its entry into the product fractionator; alternatively such treatment may be applied to the fuel oil recovered from the

fractionator, depending on process and product requirements. The former procedure may be employed when increased supplies of catalytic cracking stock are required. The latter procedure may be dictated by specification requirements of the product fuel oil, which may be affected by constituents formed in the cracking stage. Other conversion stages which may be integrated into the process of the invention include hydroforming and/or other octane improvement processes for the virgin naphtha, thermal or suspensoid viscosity-breaking of the reduced crude, etc. In accordance with the invention, however, the total effluent from at least one of these conversion stages (other than the reduced crude conversion referred to) will be supplied, together with at least one lighter fraction (if available) from the crude oil distillation stage, to the single product fractionation stage substantially as described above.

In carrying out the combined refining process described above the yield and quality of the motor fuel fraction obtained suffer to a certain extent from the fact that thermal reforming or catalytic reforming in the absence of substantial amounts of hydrogen must be used rather than catalytic hydroforming in the reforming stage to afford an efficient operation of the product fractionator. Catalytic hydroforming or similar processes are far superior to reforming processes carried out in the absence of hydrogen, with respect to product quality including the yield-octane number relationship, sulfur content and engine cleanliness of the gasoline. However, processes of this type involve a net production and recycling of large amounts of hydrogen. Introduction of this hydrogen together with the reformat into the common product fractionator would seriously interfere with the recovery of valuable hydrocarbon light ends or require a fractionator design of uneconomical dimensions.

In accordance with a feature of the combined refining process described above there may be included as one of the intermediate stages the step of catalytically reforming the naphtha fractions obtained in the various stages in the presence of hydrogen and also the use of at least part of the hydrogen produced in this reforming step to treat other suitable fractions obtained in the various stages of the combined process outside the said reformer stage and the fractionator stripper. This feature preferably involves the incorporation into the simplified combination process, of catalytic naphtha reforming operations of the type known as "Plat-

forming" wherein virgin and/or cracked naphtha from the various stages of the combined process is contacted with a platinum catalyst at elevated temperatures of about 800°—1000° F., high pressures of about 400—1000 p.s.i.g. and relatively high hydrogen recycle rates of about 3000 to 12,000 standard cu. ft./barrel maintained by recycling hydrogen produced in the process. Under the conditions specified, three major reactions occur, namely dehydrogenation of naphthenes to the corresponding aromatics, hydro-cracking of high molecular weight hydrocarbons into saturated hydrocarbons of lower molecular weight and isomerization of naphthas and straight-chain hydrocarbons. In addition, there takes place dehydrocyclization of straight-chain hydrocarbons directly into aromatics and desulfurization converting substantially all sulfur into hydrogen sulfide. Chiefly as a result of the high hydrogen partial pressure the catalyst retains its activity for many months without regeneration.

Other conventional hydroforming processes may be included in the combined process. For example, the virgin and/or cracked naphtha may be contacted in the presence of extraneous hydrogen and/or hydrogen formed *in situ* with such well known catalysts as the oxides of molybdenum, aluminum, vanadium and tungsten, with or without silica, supported on alumina, or cobalt molybdate as such or molybdena supported on zinc aluminate, at temperatures of about 800° to 1000° F., pressures of about 50 to 1000 p.s.i.g. and gas recycle rates of about 1000—3000 standard cu. ft. per bbl. of a recycle gas containing about 80% H₂. Continuous or intermittent regeneration of the catalyst may be employed, if necessary, to remove carbonaceous deposits by controlled burning with air in mixture with recycled flue gas from the regeneration process. Regeneration temperatures are generally limited to 1000°—1200° F. at any convenient pressure. The platforming or other conventional hydroforming treatment may be effected in fixed bed, moving bed, "fluid" or suspensoid type of operation, all in a manner known *per se*.

In accordance with a further feature, when incorporating such catalytic hydroforming processes into a combination process of the type specified above, hydrogen made available in the hydroforming stage is used to treat other fractions obtained in the process either with hydrogen separated from the reformer effluent or admixed therewith. Fractions which may be thus treated are light or heavy gas oils from the combined fractionator-stripper or cracked gas oil fractions which may be

- upgraded by hydrogenation. The treatment of such gas-oil fractions with hydrogen may be carried out at non-cracking conditions including temperatures of about 500° to 900° F. and pressures of about 750 to 4000 p.s.i.g. in the presence of conventional hydrogenation catalysts, such as the oxides or sulfides of Groups V, VI and VIII heavy metals. Such hydrogenation improves the quality of the gas oil stock as a catalytic cracking stock by hydrogenating coke-forming constituents such as polycyclic aromatics prior to cracking.
- The hydrogen from the hydroforming stage may also be used to upgrade product light cracked gas oil within the heating or Diesel oil boiling ranges. For this purpose the type of product oil mentioned may be hydrogenated on the catalysts of the V, VI and VIII groups just mentioned at either substantially desulfurization conditions such as temperatures of about 500°—800° F. and pressures of about 100—500 p.s.i.g., or substantially hydrogenation conditions similar to those described in the preceding paragraph. Rather than using separated hydrogen, the total hydroformer effluent may be used to upgrade the light cracked product gas oil as will appear more clearly hereinafter.
- The reformed naphtha may be blended with the naphtha fractions recovered from the fractionator-stripper, preferably after the latter have been subjected to conventional finishing treatments such as caustic washing and/or suitable sweetening processes known to the art. The reformed naphtha itself is of such high quality and purity as not to require further finishing of this character.
- While these types of operation may require special gas-liquid separation and/or fractionation means in addition to the combined fractionator-stripper, the resulting increased investment is at least in part compensated by the improved quality of the final motor fuel blend. Furthermore, the naphtha finishing equipment may be substantially reduced in size whereby appreciable savings in investment are secured.
- In accordance with another feature, the total effluent of the catalytic naphtha hydroforming stage including all the hydrogen present in this stage is used without intermediate cooling to strip at least a portion of the reduced crude prior to its entry into the combined fractionator-stripper in a separate stripping stage operated substantially at the pressure of the hydroforming stage. Assuming a temperature of the reduced crude recovered from the crude still of about 600° to 750° F. and a temperature of about 800°—1000° F. of the hydroformer effluent this preliminary stripping may take place over a temperature range of about 650° to 800° F. at a pressure of about 100 to 1000 p.s.i.g. At these conditions, a substantial proportion of the gas-oil and lighter constituents of the reduced crude, say about 5 to 50%, will go overhead with the hydroformate and hydrogen. Hydrogen may then be separated from normally liquid overhead and recycled to the hydroforming stage after suitable reheating. The hydroformate containing the reduced crude strippings may be passed to the fractionating section of the combined fractionator-stripper or fractionated in a separate tower to produce gas oil to be used as catalytic cracking stock and product hydroformate which may be blended with the finished naphtha from the combined fractionator-stripper as described above. The pre-stripped reduced crude is passed to the stripping section of the combined fractionation-stripper to be further stripped therein with the effluent vapors of other process stages.
- In the embodiments described above wherein a hydroforming stage is included, excess hydrogen made available in the hydroforming stage beyond the recycle requirements of this stage may be passed to the stripping section of the combined fractionator-stripper and together with virgin and cracked naphtha or any suitable mixture of such heavy naphthas. It may be desirable to make all virgin naphtha available for stripping the reduced crude in the combined fractionator-stripper. In this event a combined virgin light and heavy naphtha stream may be passed directly to the stripping section of the fractionator-stripper and only a heavy naphtha fraction from the fractionator-stripper may be subjected to catalytic hydroforming as described.
- Various embodiments of the invention will now be described with reference to the accompanying drawings wherein:—
- Figure 1 and continuation Figure 1a show a schematic flow plan of one combination process according to the invention;
- Figure 2 and continuation Figure 2a show a schematic flow plan of a modified combination process according to the invention, which includes stages for hydroforming one or more naphtha streams;
- Figure 3 shows one form of hydroforming stage; and
- Figure 4 shows a modification of the arrangements shown in Figures 1 and 1a, and 2 and 2a.

In the figures similar parts have been indicated by the same reference number.

Referring now in detail to Figures 1 and 1a, of the drawings, the system illustrated therein essentially comprises a crude still 110, a naphtha reformer or hydroformer schematically shown by element 130, a product fractionator 140, a cracking stage schematically illustrated by element 150, and fuel oil filtering facilities at 190. The functions and coaction of these elements will be forthwith explained using as an example the refining of a medium gravity crude of the type of Arabian Qatar crude in a refinery having a capacity of about 10,000 bbl. of crude per day. It should be understood, however, that the system may be used for the refining of different types of crude at a larger or smaller scale in a generally analogous manner.

In operation, the crude oil is pumped from line 101 by means of pump 103 via line 105 through heat exchangers 106 to a heating coil located in furnace 107 wherein it is heated to a temperature suitable to vaporize a substantial portion of the oil. The oil so heated is passed through line 109 to a lower portion of still 110 which it may enter at a temperature of about 750°—800° F. and a pressure of about 40—70 p.s.i.g. Still 110 is provided with a plurality of horizontal bubble cap plates 112 to improve fractionation of the feed in a conventional manner. Reflux may be accomplished with the aid of partial condensers 114 arranged internally or externally at the top of still 110. For the purposes of the present example, still 110 may be so operated that three distillate streams and distillation bottoms are produced as follows.

All crude constituents boiling below about 250° F. are removed together as a vapor stream of light virgin naphtha overhead through line 116 at a temperature of about 300°—350° F. This stream may amount to about 20 to 25% of the crude charged. A liquid stream of heavy naphtha having a boiling range of about 250°—400° or 500° F. is removed through line 118 from an upper portion of still 110 at a point below condensers 114. About 20 to 25% of the crude charged is recovered through line 118. A kerosene or Diesel oil boiling within the range of about 400° to 700° F. and amounting to about 17 to 23% of the crude is drawn off through line 120. The remainder of the charge, amounting to about 40 to 50% and consisting predominantly of constituents boiling above 700°—800° F. is withdrawn as reduced crude through line 122 from the bottom of still 110. The

kerosene removed through line 120 is normally suitable for kerosene or Diesel oil purposes without further treatment and it may be passed directly to storage. The other fractions may be treated in accordance with the present invention as will be forthwith described.

The light virgin naphtha vapors in line 116 may be passed directly to a lower portion of product fractionator 140. If desired, this vapor stream may be preheated to about 800° to 1000° F. to conform with the heat requirements of fractionator 140. This may be done by bypassing at least a portion of the vapors in line 116 through a heating coil 117 either separate or as shown in Fig. 1 located in convection section 132 of reformer furnace 130 operated as will be described hereinafter.

The heavy naphtha stream in line 118 may be pumped by pump 126 through line 128 at a pressure of about 900—1100 p.s.i.g. to a thermal reforming or catalytic hydroforming stage. This reforming stage may be of any conventional design well known in the art, and may comprise a conventional tube furnace 130 designed to provide for an oil residence time of about 18 to 25 liquid volumes per volume of reaction space per hour ($v/v/hr.$) at about 1000°—1100° F. and about 1000 p.s.i.g. pressure. Under these conditions, the octane rating of the naphtha may be increased from about 20 to 30 to about 70 to 80 Research Octane Number without excessive cracking to normally gaseous hydrocarbons. The total effluent from reformer 130 is passed substantially at the reforming temperature and pressure through line 134 provided with a pressure release device such as valve or venturi 136 into a lower portion of fractionator 140 at a point close to the feed point of line 116. As a result of the pressure release to fractionator pressure of about 5—15 p.s.i.g., the naphtha is substantially completely vaporized when entering the lower portion of fractionator 140.

The reduced crude in line 122 may be passed directly to the top of the lower portion of fractionator 140, substantially at the temperature of its withdrawal from still 110. Line 122 feeds into fractionator 140 at a point above the feed points of lines 116 and 134. In this manner, the vapors supplied through lines 116 and 134 pass upwardly through fractionator 140 against the downwardly flowing reduced crude to strip the latter of vaporizable constituents. This effect and the operation of fractionator 140 will be described in greater detail later on.

At this point it is noted that a side stream of gas oil range hydrocarbons 130

amounting to about 45 to 60% on crude and having a boiling range which renders it suitable as a catalytic cracking stock, may be withdrawn from an intermediate section of fractionator 140 via a gas oil reflux system comprising pump 143 and lines 142, 144 and passed through line 146 to a catalytic cracking stage 150. This cracking stage is preferably carried out as described in our copending Application No. 5650/51, namely a two vessel cracking and catalyst regeneration system operating with beds of finely divided fluidized solid materials in each of the vessels and continuous circulation of the solid materials between the two vessels by means of two U-shaped conduits depending below the said vessels and each having one of its vertical legs opening into and communicating with the bed in each the said vessels and in which the circulation of the solid material is controlled by differential pressures in the vertical legs of the U-shaped conduits without the use of mechanical valves. This is a preferred embodiment. However any other conventional cracking system adapted to convert gas oil range hydrocarbons into lower boiling oils, particularly of the motor fuel range, may be used. Continuous or batch operation may be employed in fixed bed, moving bed, "fluid" or suspension systems. Heat required for cracking may be supplied as preheat of process materials and/or as sensible heat of exothermically regenerated catalyst or in any other conventional manner. Modified natural or synthetic clay or gel type catalysts such as activated montmorillonite clays, silica-alumina, silica-magnesia composites and other conventional cracking catalysts may be employed at temperatures of about 800°—1000° F. and pressures of about 45 atmospheric to 25 p.s.i.g., all in a manner known *per se*.

The total hydrocarbon effluent of cracking stage 150 is passed substantially at the cracking temperature of, say, about 800°—900° F. through line 152 to the lower portion of fractionator 140, preferably at a point intermediate between the feed points of the reduced crude on the one side and of the reformed and virgin naphthas on the other side. If cracking stage 150 is operated at an elevated pressure, the pressure may be released by valve 152a to fractionator pressure. In most cases, about 95 to 100% of the cracked material enters fractionator 140 in the vapor state to enhance the stripping action of the vapors supplied through lines 116 and 134, while any unvaporized constituents of the cracked

material are in turn subjected to stripping by those vapors introduced through lines 116 and 134.

As indicated in Figures 1a and 2a, fractionator 140 comprises a lower stripping section A and an upper combined fractionation-absorption section B. Both sections are provided with suitable means for improving the countercurrent contact between downflowing liquid and upwardly flowing vapors. For the purpose of stripping, a disc-and-doughnut baffle arrangement has been found to be most efficient and such is shown schematically for section A by elements 141. Section B is illustrated to contain a number of bubble cup plates 154 to enhance the efficiency of the fractionation-absorption process. Sections A and B may operate as follows.

Stripping section A receives, in addition to the vapour and liquid streams supplied through lines 116, 134, 152 and 122, a liquid top feed comprising a gas oil cut removed from the bottom of section B via line 142 and supplied to section A via line 144. This gas oil is fed to section A to provide control over the reflux and heat removal in that section in order to obtain the desired end point and clean up on the gas oil. All the heat required for stripping and fractionation in fractionator 140 is preferably supplied as sensible heat of the hydrocarbon streams entering section A to maintain a temperature of, say, about 820°—830° F. in the lowest portion of section A. The vapors rising through section A strip the downwardly flowing gas oil, reduced crude and cracked liquid products of substantially all their distillable constituents and this vapor mixture passes on at a temperature of about 700° to 750° F. into fractionation-absorption section B to be treated as will be described later on.

The reduced crude from line 122 which may contain as much as about 75% of gas oil suitable for feed to the catalytic unit is countercurrently stripped and heated by the cracked vapours at, say, about 875° and 8 p.s.i.g. and then by the virgin naphtha at, say, about 800° F. and by the reformat at about 1025° F. The partial pressure effect of the stripping streams and the heat content thereof are sufficient to cause the gas oil constituents of the reduced crude to vaporize. The net effect of the process in section A then is (1) a bottom stream of unfluxed fuel oil amounting to about 10—15% on crude and containing about 85—90% of flashed reduced crude, about 8—9% of reformer tar, and about 0.3% of heavy slurry oil from the catalytic operation, all blended automatically so that it may be fluxed with about 50% of light Diesel oil blend-

ing stock for fuel oil viscosity correction; (2) vapours containing all of the distillate products to be obtained from fractionator 140 and leaving section A overhead at about 800° F.

A heavy material containing all the non-distillable constituents of the crude charged and of the fractions converted in stages 130 and 150 collects at about 820°—830° F. in the bottom zone of section A from which it may be withdrawn via line 147. If desired, the temperature in the bottom of section A may be reduced to, say, about 700° F. by recycling heavy bottoms from line 147 by means of pump 149 via line 151, cooler 153 and line 155. The bottoms quenching may be desirable to prevent cracking and coking of the heavy liquid products. Combined reduced crude amounting to about 12 to 18% on crude may be recovered through line 181 to be further treated as will appear hereinafter.

At the conditions of the present example about 1000 to 15000 mols./hr. of hydrocarbon vapors will be available to strip in section A about 100 to 200 mols./hr. of liquid. This favorable vapor-liquid ratio results in a substantially quantitative stripping effect being afforded in section A. The number of baffles and the dimensions of section A depend largely on the character of the crude charged and the products desired. For the purposes of the present example, this section may be approximately 12' in diameter and 30' in height and is equipped with 7 sets of disc-and doughnut-type contacting devices.

Vapors passing upwardly to section B of the fractionator 140 are fractionated with gas oil reflux through line 144 and cooler 145. Gas oil may be withdrawn through line 142 at about 650°—750° F. This gas oil contains all of the virgin 600°—1050° F. cut point fraction and all of the cycle gas oil, and the gas oil is hence run to extinction. Gas oil withdrawn from line 142 may be divided into 3 streams, namely (1) reflux through line 144 as described; (2) a liquid product amounting to about 2.5—3% on crude for heavy Diesel oil blending through line 148 and cooler 161 to tankage; (3) about 50—55% on crude of catalytic feed stock which is fed directly to the cracking section by pump 143 through line 146.

Passing now to section B of fractionator 140, it is noted that in most conventional catalytic cracking and similar refining operations, products are fractionated at low pressures to produce a gas and low pressure distillate. The gas stream contains appreciable quantities of gasoline constituents and it is, therefore, necessary to compress, absorb, and refrac-

tionate this stream to recover its gasoline constituents. This may be avoided by combining both low pressure adsorption and fractionation in the upper section A of fractionator 140. For this purpose, one of the "pumparounds" normally used merely for heat removal and returned to a point close to its withdrawal may be used as an absorption medium by returning it to a point substantially above that from which it is withdrawn.

The operation of the upper part of section B will be briefly described herein insofar as it contributes to the essential advantage of this invention, which is directed towards making small refineries fully competitive with large refineries.

Product vapors leaving the gas oil fractionating section and entering the heating oil withdrawal section contain heating oil, total naphtha and gas. In the section immediately above the heating oil withdrawal line 156, these product vapors which may be at about 510°—520° F. are cooled by contact with cool heating oil at about 130° F. entering the tower through line 168. Heating oil is condensed out and falls along with the cooled part entering through line 168 and both are withdrawn from the tower through line 156. The cool heating oil in line 168 is saturated with C_5 , C_6 and C_7 homologues to form a fat oil. The light fractions are stripped out by ascending naphtha and gas product vapors, thus increasing the concentration of said fractions in the naphtha condensing zone.

Heating oil through line 156 is withdrawn at about 480°—500° F. and most of it is cooled in cooler 158 and returned to the top of section B by pump 160 and line 162 as absorber lean oil, the remainder being stripped in stripper 157 and taken through line 178 to product or as a fluxing oil to the heavy residue in the bottom of the tower via lines 183, 181, and 155, vapors being returned through line 159.

Product vapors entering the portion of section B which lies above the inlet of line 168 thus consist of the naphtha fractions and gas fractions normally encountered plus an abnormal quantity of C_4 , C_5 and C_6 which were absorbed as previously explained in the top of section B. Such vapors and heating oil are fractionated by reflux pumped back through line 176 and the fractionated vapors may be taken from the tower at about 5 p.s.i.g. and 215° F. through line 166 and may be compressed by a one stage blower 169. Naphtha with an excess of light fractions is condensed in condenser 172, the excess light fractions thereof being removed in separator 173 and, except for the part returned as reflux through line 176, sent

to final product tankage through line 170.

Gas and light fractions are led back to the absorber portion of section B through line 177 where countercurrent absorption of the desired light fraction by the lean oil as previously described takes place. The number of plates between lines 156 and 162 is preferably increased by about 10 to 15 over that of normal fractionator designs. By using this technique it is possible to absorb essentially all of the C_5 + fractions in the gas entering the top section of fractionator 140. In addition, as much as 75% of the C_4 components can be absorbed. The gas leaving the top of fractionator 140 through line 164 is thus stripped of its valuable gasoline components and can be passed directly to fuel uses. A naphtha or gasoline cut may be recovered via line 166. Such gasoline can be withdrawn at temperatures of about 120°—150° F in spite of the low pressure employed.

The gasoline fraction is uncontaminated with the "pump-around" medium as described above and is condensed either in the tower by naphtha pump-back or in an external condenser, part of the naphtha being returned to the tower for fractionation. The number of plates to be provided in section B depends on the type of crude charged and the products desired. For the purposes of the present example, 2 plates may be used between the reduced crude inlet and gas oil withdrawal, 4 between gas oil and heating oil, 5 for stripping of gas fractions from the heating oil, 3 for naphtha-heating oil fractionations, and 10—15 plates for the absorption of light components in the top portion of section B.

As indicated in the drawing, final products may be recovered from the fractionator 140 as follows. Gasoline of 400° F. end point amounting to about 50 to 60% on crude and having an octane rating of about 75 to 85 may be passed via line 170 to tankage. About 20 to 30% of gasoline on crude may be recirculated by pump 174 via line 176 to section B to serve as reflux. Final heating or light Diesel oil may be recovered via line 178 at a rate of about 0.5 to 1.0% on crude. About 2—3% on crude of a heavy Diesel oil stock may be obtained via line 148.

Returning now to the combined reduced crude type bottoms withdrawn through line 181, they may, if desired, be blended with gas oil or lighter fractions supplied through line 183 to adjust their viscosity to meet specifications. The bottoms may then be cooled to about 200° to 500° F. in cooler 185 and passed through line 187 to filtering facilities 190. Conventional sand filters, rotary or porous sintered

ceramic filters may be used to remove from the combined residue all suspended or slurried solid particles, such as coke, catalyst carried over from cracking stage 150, etc. The solids removed in filtering facilities 190 may be discarded via line 192 or passed to catalyst recovery means (not shown). A fuel oil grade residuum is recovered via line 194. Combination filtering in this manner of thermally cracked tar, catalytic slurry and crude residuum avoids "compatibility" problems arising upon conventional blending of materials of this type and permits the recovery of a fuel oil relatively low in sediment.

The system illustrated in Figure 1 permits of various modification. As previously pointed out, one or more reduced crude viscosity breaking or coking stages may be included. Certain of the effects contemplated by such coking stages may be accomplished by operating the bottom portion of section A of fractionator 140 at viscosity-breaking conditions, for example at 875° to 950° F. Instead of, or in addition to, this type of operation, the reduced crude in line 122 may be subjected to a conventional thermal or catalytic coking or viscosity-breaking treatment. The total effluent of such a coking stage, with or without the coke formed may be passed on through line 122 to an upper portion of stripping section A of fractionator 140 to be treated therein as above described. Similarly, part or all of the bottoms in line 147 may be subjected, for example, in unit 182 to viscosity-breaking and/or coking to produce a heavy residuum and coke to be passed on from a separator 184 through lines 186 and 181 to filtering means 190 and lighter materials which may be returned from separator 184 through line 188 to a middle portion of section A to be subjected therein to stripping and fractionation as described above.

Catalytic, rather than thermal, reforming may be employed in reforming stage 130 using such conventional catalysts as oxides and sulfides of Group V, VI, or VIII metals, preferably supported on a suitable carrier, temperatures of about 850°—1100° F. and pressures from atmospheric to about 400 p.s.i.g. in the presence or absence of extraneous hydrogen, all in a manner known *per se*. Other refining treatments, such as bauxite treating or clay treating, may follow reforming stage 130, provided that most of the hydrocarbon effluent of such stages is supplied to fractionator 140 as described above.

While a variety of catalytic cracking systems may be used as cracking stage 130

stantially all sulphur is removed. The hydrogenated gas oil may then be passed on through line 146 as described above.

5 In the embodiment shown in Figs. 2 and 2a of the drawing, final products may be recovered from tower 140 in a slightly different manner from that shown in Figs. 1 and 1a above as follows. Gasoline of 10 400° F. end point amounting to about 35 to 45% on crude and having an octane rating of about 80—90 Research may be passed via line 170 to a conventional finishing stage 221 and then to tank 222 15 to be blended therein with the hydroformate supplied via line 220. Gasoline may be recirculated by pump 174 via line 176 to section B to serve as reflux. Final heating or light Diesel oil may be recovered via 20 line 178 at a rate of about 0.5—1.0% on crude. About 2—3% on crude of a heavy Diesel oil stock may be obtained via line 148. A naphtha fraction somewhat heavier than that of the gasoline in line 166 and having a boiling range of, say, 25 about 200 to 400° F. may be withdrawn from a point intermediate between the withdrawal point of line 156 and the feed point of line 168 to be passed via line 129 30 to hydroformer 130 as described above. The amount of naphtha so withdrawn may vary between 0 and about 45% on crude depending on the proportion of virgin naphtha directly supplied to hydro- 35 former 130 via line 128.

Another embodiment of the invention is illustrated in Fig. 3 which, for the sake of simplicity, shows only those 40 elements which differ in design and/or operation from those of Figs. 2 and 2a and in addition some of the pipe lines of Figs. 2 and 2a to clarify the manner 45 in which the equipment illustrated in Fig. 3 is to be linked up with the combination unit described with reference to Figs. 2 and 2a. In essence, the elements shown in Fig. 3 are intended to take the place of hydroformer 130, stabilizer 200 50 and hydrogenators 216 and 218 of Figs. 2 and 2a.

Referring now in detail to Fig. 3, the system illustrated therein essentially comprises three catalytic hydroformers 300, 302 and 304 connected in series, a hydro- 55 fining stage 320 and a stabilizer-fractionator 330. The function of these elements in a combination unit of the type illustrated in Figs. 1, 1a, 2 and 2a will be forthwith described with reference to 60 a "platforming" operation taking place in hydroformers 300, 302 and 304. It should be understood, however, that other types of catalytic hydroforming may be carried out in this system in a substantially 65 ally analogous manner.

In operation, naphtha having a boiling range of about 150° to 400° F. may be passed through line 128 to hydroformer 300 via reheater 307 at a temperature of about 900° to 1000° F., a pressure of 7 about 50 to 1000 p.s.i.g. and in an amount of, say, about 15 to 45% on crude. This naphtha may be supplied 7 from crude still 110 via line 118 and/or from tower 140 via line 129 in any desired proportion, as described with reference to Figs. 1, 1a, 2 and 2a. An amount of about 8 500 to 7000 SCF/Bbl. of recycle hydrogen is supplied to hydroformer 300 from line 306 at hydroforming pressure and a temperature of about 900° to 1200° F. Hydro- 8 formers 300, 302 and 304 contain a platinum group catalyst such as platinum or palladium supported on alumina pellets preferably arranged in fixed beds. The 9 naphtha-hydrogen charge passes in series through hydroformers 300, 302 and 304 and is reheated between stages in 9 reheaters 308 and 310 to maintain similar temperatures in the three stages. Suitable "platforming" conditions to be main- 9 tained in hydroformers 300, 302 and 304 include temperatures of about 800° to 1000° F., total pressures of about 400 to 1000 p.s.i.g. and hydrogen recycle rates 9 of about 3000 to 12000 SCF/Bbl. At these conditions the octane number of the naphtha is improved anywhere from 30— 60 points and the sulfur removed almost completely. The catalyst retains its 1 activity for many months whereupon it may be replaced with fresh or regenerated catalyst, no continuous or frequent 1 periodic regeneration being required.

The total effluent from the hydroform- 1 ing or "platforming" stages passes through line 312 to hydrofining stage 320. Simultaneously, hydrofining stage 320 1 receives via line 119 all or any desired portion of the light gas oil range product 1 fraction recovered from tower 140 prior to hydrogenation in hydrogenator 216. The combined charge is treated in hydro- 1 fining stage 320 with a hydrofining catalyst such as oxides or sulfides of Group 1 V, VI and VIII metals at temperatures of about 500° to 800° F., pressure of about 50 to 1000 p.s.i.g. and throughputs of 1 about 0.5 to 5 V/V/Hr. At these conditions substantially complete conversion of 1 undesirable constituents of the light gas oil fraction, such as sulfur compounds, takes place by straight hydrogenation or 1 hydrogen transfer from the platformate, without appreciably effecting the quality 1 of the platformate. As a result of the high hydrogen partial pressure, catalyst 1 deactivation by carbon deposition is relatively slow, eliminating the requirement of continuous or frequent periodic 1

150. "fluid" catalytic cracking involving the continuous production of cracked effluent and continuous catalyst circulation between cracking and regeneration stages described in our co-pending Application No. 5650/51 is particularly suitable.

Referring to Figs. 2 and 2a the embodiment illustrated is closely similar to that described above and illustrated in Figs. 1 and 1a. The operation of the system is substantially the same as described in connection with Figs. 1 and 1a with the principal modification that the catalytic or thermal reforming stage 130 is replaced by a catalytic hydroforming unit.

In operation of this unit the heavy naphtha stream may be pumped by pump 126 through line 128 at a pressure of about 600—900 p.s.i.g. to a catalytic hydroforming stage 130. This hydroforming stage may be of any conventional design well known in the art and is preferably one suitable for a "platforming" operation or a regenerative hydroforming operation of the type specified above. For the present example, a conventional catalytic reforming operation carried out at about 800° to 1000° F., 50 to 1000 p.s.i.g. pressure nominal oil residence time of about 0.5 to 2.5 hours calculated as cold oil in contact with catalyst, and hydrogen recycle rates of about 500 to 7000 SCF/Bbl. of oil using a catalyst, such as molybdena supported on alumina, in fixed bed operation is referred to. In addition to the virgin naphtha supplied to hydroforming stage 130 via line 128, a hydrocarbon stream of similar boiling range derived from the fractionator section of tower 140 may be fed to hydroforming stage 130 via line 129 in any desired proportion. Under the conditions specified the octane rating of the naphtha may be increased from about 25 to 40 to about 80 to 100 Research Octane Number without excessive cracking to normally gaseous hydrocarbons and carbon. A plurality of hydroforming reactors may be provided alternating between onstream and regenerating cycles, oxidizing gases such as air, steam, flue gases or mixtures thereof, being used in the regenerating cycle to remove carbon from the catalyst all in a manner known *per se*. Of course, any type of continuous operation such as fluid moving bed or suspensoid operation may be used wherein the catalyst is continuously circulated between reactor and regenerator vessels.

The total effluent of hydroformer 130 is passed through line 134 via heat exchanger 117 to a stabilizer tower 200 wherein hydrogen and normally gaseous hydrocarbons such as C₁—C₃ hydrocar-

bons are separated from normally liquid hydroformate at temperatures of about 80° to 120° F. and pressures of about 100 to 1000 p.s.i.g.

Gases containing about 40 to 90% of hydrogen and amounting to about 1000 to 9000 SCF/Bbl. of feed are withdrawn from stabilizer 200 through line 202 and passed through recycle booster 204 and reheater 206. An amount corresponding to about 500 to 7000 SCF of hydrogen per bbl. of feed may be passed via line 208 to hydroformer 130 substantially at the temperature and pressure of the latter. The remainder of the gases in line 202 or any desired portion thereof may be passed through line 210 via booster 211 and lines 212 and/or 214 to hydrogenation stages 216 and/or 218 which may be used to improve the cracking characteristics of cracking recycle stock supplied from tower 140 through line 142 or to improve the quality of product light gas oil recovered from tower 140 via line 178 as will appear more clearly hereinafter. Any excess gases may be withdrawn through line 203 and passed through line 203a to the bottom portion of fractionator-stripper 140 as an additional stripping agent or used as fuel gases in the system. When operating in this manner for example, the gravity of recycle gas oil can be increased 2—15 degrees, and the stock made as desirable as is virgin feed for catalytic cracking.

The hydroformate separated in stabilizer 200 is withdrawn through line 220. This material may be blended without further treatment in tank 222 with finished gasoline recovered from tower 140 as will appear more clearly hereinafter. The hydroformate so recovered may amount to about 10 to 25% on crude.

As shown in Fig. 2a part of the side stream of gas oil hydrocarbons removed from the fractionator via pump 143 and lines 142 may be passed through a hydrogenation stage 216.

For example, all the heavy gas oil not to be returned as reflux to the fractionator 140 via line 144 may be supplied by pump 143 to hydrogenation stage 216 and hydrogenated therein with nickel tungsten sulfide as the catalyst at a temperature of about 750° to 800° F., a pressure of about 2500 to 3000 p.s.i.g., a liquid throughput of about 1 to 2 V/V/Hr. and a hydrogen feed rate of about 2500 to 3000 standard cu. ft. per bbl. of oil in fixed bed operation. Under these conditions about 1000—1200 standard cu. ft. of hydrogen per bbl. is consumed in the hydrogenation operation. The API gravity of the heavy cycle gas oil is increased by about 10—12 units and sub-

regeneration. Fixed bed operation is, therefore, suitable for hydrofining stage 320. However, any conventional continuous or periodic regeneration system involving fluid or moving bed operation may be employed if desired.

The effluent of hydrofining stage 320 is passed through line 322 to an intermediate section of conventional stabilizer-fractionator 330. This tower is so operated that a bottoms fractions of high quality heating or Diesel oil may be recovered and passed to tankage via line 332, platformate having a final boiling point of about 400° F. may be passed via line 334 to blending tank 222 and gases containing about 80 to 95% of hydrogen and amounting to about 5000 to 12000 SCF per Bbl. are recovered overhead through line 336. About 500 to 1500 SCF per Bbl. of the gas in line 336 may be vented as excess. The remainder is recycled by means of a recycle booster 340 via lines 338 and 342 and reheater 344 to line 306 and hydroformer 300.

Rather than supplying merely the product light gas oil fraction from line 119 to hydrofining stage 320, any desired portion of the recycle stock from line 146 may be used for this purpose. In this case, the oil fraction recovered through line 332 may be returned as a whole or in part to line 146 to serve as an upgraded recycle stock for cracking stage 150.

In all other respects, operation of the entire combination unit may be substantially as described with reference to Figs. 1 and 1a, including operation of crude still 110, tower 140, cracking stage 150, coking stage 182, etc.

A still further embodiment of the invention is illustrated in Fig. 4 which is a schematical flow plan of a complete combination unit similar in various respects to that shown in Figs. 1, 1a, 2 and 2a. corresponding elements, though illustrated in a simplified manner, being as before identified by like reference numerals. An important additional element is a high pressure stripper 435 whose function will be forthwith described.

Referring now in detail to Fig. 4, crude oil is charged via line 109 to crude still 110 and treated therein substantially as described with reference to Figs. 1 and 2. A kerosene or light gas oil fraction ready for use may be recovered via line 120. Light naphtha is withdrawn through line 116, heavy naphtha through line 118 and reduced crude through line 122. In addition, a gas oil fraction may be recovered through line 420 and passed wholly or in part to cracking stage 150 to serve as virgin cracking stock therein. Virgin gas

oil may be recovered as product via line 421. Any desired proportion of the light and heavy naphthas may be recovered via lines 416 and 418 as product or passed directly to the bottom of tower 140 as described with reference to Figs. 1—3.

In accordance with a preferred feature of the embodiment under discussion the heavy naphtha from still 110, amounting to about 15 to 20% on crude, is passed to line 128 and hydroforming stage 130. The latter may be operated substantially as described for stage 130 and hydroformers 300, 302 and 304 with reference to Figs. 1 and 3, recycle hydrogen being added via lines 406 as will appear more closely hereinafter.

The total effluent of hydroforming stage 130 is passed substantially at the temperature and pressure of stage 130, say, at about 800°—1000° F. and about 700—800 p.s.i.g. through line 432 to a bottom portion of high-pressure stripper 435. This stripper which may be of conventional design, having a lower disc-and-annulus section C and an upper bubble tray section D, is maintained substantially at the pressure of reforming stage 130.

Reducer crude from line 122 is passed by pump 437 through line 439 at the pressure of stripper 435 to the top of section C of stripper 435 which is so designed that the temperature varies from about 600° to 800° F. in the bottom to about 500° to 750° F. in the top of stripper 435. At these conditions, the reduced crude supplied through line 439 is stripped of its gas oil and lower boiling constituents to an extent of about 5—35% on reduced crude depending on the type of crude involved. The stripped reduced crude now amounting to about 25 to 45% on crude is withdrawn through line 441 provided with pressure release valve 443 and passed to an upper portion of section A of tower 140, to be further stripped therein by cracked vapors supplied through line 152 from cracking stage 150 substantially as described with reference to Figs. 1—2a.

A combined overhead containing the reduced crude strippings consisting chiefly of gas oil, the total hydroformate and the total hydrogen is withdrawn from stripper 435 via line 445, cooled in cooler 447 to about 80° to 120° F. and passed to gas liquid separator 449 substantially without pressure reduction other than that caused by cooling. The gas separated in separator 449, amounting to about 500 to 5000 SCF/Bbl. and containing about 40 to 95% of hydrogen is withdrawn through line 406, reheated in reheater 451 to about 900° to 1200° F. and returned to hydroforming stage 130. Excess gas may be released through line 130

407 to be used as a fuel gas or in tower 140 as an additional stripping medium.

The total liquid separated in separator 449 may be passed through lines 453 and 455 provided with pressure release valve 457 to the fractionating section B of tower 140 to be fractionated therein substantially as described with reference to Fig. 2a. If desired, part or all of the liquid in line 453 may be passed to a separate fractionator 460 after pressure release in valve 459. High quality hydroformate product may be recovered overhead from tower 460 and recovered via line 462 to be blended in tank 222 of Figure 2a with finished gasoline from tower 140. A gas oil fraction may be withdrawn from tower 460 at bottoms via line 464 to be combined with the gas oil in line 142 and further treated as described with reference to Figs. 1 and 1a.

Similarly as explained with reference to Figs. 2, 2a and 3, the virgin naphtha in lines 116 and/or 118 may be in part supplied directly to the bottom of tower 140 to serve as a stripping medium for the reduced crude and corresponding amounts of similar fractions from tower 140 may be used as charging stock for hydroformer 130, as will be readily understood by those skilled in the art. Also, any desired portion of the virgin light naphtha in line 116 may be passed to reforming stage 130.

What we claim is:—

1. A combination process for the refining of petroleum oils which comprises distilling a crude petroleum oil to yield a residual fraction and a plurality of lighter fractions, subjecting at least one of said lighter fraction to a conversion treatment, passing said residual fraction to a fractionating tower and feeding a plurality of hydrocarbon streams into said fractionating tower below the point of entry thereto of said residual fraction whereby the latter is stripped by the vapors of said hydrocarbon streams, and recovering the desired product fractions from said fractionating tower, wherein said hydrocarbon streams fed to the fractionating tower comprise at least one total effluent from a conversion treatment of said lighter fractions and, if available, at least one of said lighter fractions as such.

2. A process as claimed in Claim 1, wherein at least one of the product fractions from said fractionating tower is subjected to a conversion treatment and the products from said conversion treatment are fed to the said fractionating tower below the point of entry of said residual fraction.

3. A process as claimed in Claim 2, wherein said product fraction is a gas oil fraction, which is subjected to a catalytic cracking treatment.

4. A process as claimed in Claim 3, wherein said gas oil fraction is mixed with a virgin gas oil fraction from the crude oil distillation stage, the mixture being subjected to the cracking treatment.

5. A process as claimed in Claim 3 or 4, wherein the cracking of the gas oil is carried out by employing a method or apparatus as claimed in any of the claims of co-pending Application No. 5650/51 (Serial No. 716,242).

6. A process as claimed in any one of Claims 1—5, wherein a naphtha fraction from the crude oil distillation stage is subjected to a thermal or catalytic reforming treatment and the reformed product is fed to said fractionating tower.

7. A process as claimed in any one of the preceding claims wherein the bottoms fraction from said fractionating tower is filtered and passed to storage.

8. A process as claimed in any one of the preceding claims wherein a naphtha fraction derived from the combination process is subjected to catalytic reforming treatment yielding hydrogen.

9. A process as claimed in Claim 8, wherein the said catalytic reforming treatment comprises treating the naphtha fraction in the presence of a platinum group catalyst at a temperature between 800 and 1000° F. under a pressure between 400 and 1000 lbs. per sq. in. with a hydrogen feed of between 3000 and 12000 standard cubic feet per barrel of naphtha.

10. A process as claimed in Claim 8 wherein the catalyst contains a catalytically active metal oxide such as molybdenum, vanadium or tungsten oxide.

11. A process as claimed in any one of Claims 8—10 wherein the naphtha fraction submitted to the catalytic reforming treatment is a naphtha fraction from the crude oil distillation stage or a mixture thereof with a naphtha fraction from said fractionating tower.

12. A process as claimed in Claim 11 wherein the said naphtha fraction is heavy naphtha fraction.

13. A process as claimed in any one of Claims 8—12 wherein at least one fraction obtained from the crude oil distillation stage or said fractionating tower is treated with the hydrogen from the catalytic reforming stage externally of said crude oil distillation stage and fractionating tower.

14. A process as claimed in Claim 13 wherein the treatment of said fraction with hydrogen comprises treatment of 130

said fraction with the total effluent from the catalytic reforming stage.

15. A process as claimed in Claim 13 or 14 wherein said fraction treated with
5 hydrogen is a light gas oil fraction or a heavy gas oil fraction withdrawn from said fractionating tower.

16. A process as claimed in Claim 15, wherein the treatment of the gas oil frac-
10 tion with hydrogen is followed by a catalytic cracking treatment.

17. A process as claimed in Claim 13 or 14, wherein a gas oil fraction is cata-
15 lytically cracked and the cracked products are treated with hydrogen from the reforming stage.

18. A process as claimed in any one of Claims 8—17, wherein a major proportion
20 of the heat required in said fractionating tower is supplied as the sensible heat of the products of the various conversion stages supplied thereto.

19. A process as claimed in any one of the preceding claims, wherein said frac-
25 tionating tower has a lower stripping section and an upper combined fractionation and absorption section, said residual fraction is supplied to the top of said stripping section, said plurality of hydro-
30 carbon streams are supplied to a lower portion of said stripping section, product distillate and gas oil fractions are removed from said upper fractionation section, and residual fraction bottoms are
35 recovered from the bottom of said lower stripping section.

20. A process as claimed in Claim 19, in which said residual fraction is stripped
40 by subjecting it to a two-stage heating, vaporization and stripping action, first with vapors of the effluent from a catalytic cracking treatment of a gas oil fraction, and secondly with the vapors of other fractions and conversion products
45 derived from the process.

21. A process as claimed in Claim 20, in which the vapors of cracked gas oil
50 fraction are themselves stripped in said fractionating tower with vapors of other fractions or conversion products supplied thereto.

22. A process as claimed in Claim 21, in which at least a portion of said cracked gas oil fraction vapors is heavier but of

a lower temperature than said vapors of 55 the other fraction or conversion products.

23. A process as claimed in any one of the preceding claims in which a fraction of the fuel oil boiling range, other than
60 said residual fraction, obtained from said crude oil is subjected to a high temperature thermal treatment at conditions severe enough to cause at least a reduction of viscosity.

24. A process as claimed in Claim 23 in
65 which vapors produced in said thermal treatment are passed to a lower portion of said fractionating tower wherein said stripping takes place.

25. A modification of the process 70 claimed in any one of the preceding claims wherein the residual fraction from the crude oil distillation stage instead of being passed as such to said fractionating tower is first treated with hydrogen to
75 strip lighter fractions therefrom.

26. A modification according to Claim 25, wherein said residual fraction is stripped with a hydrogen-containing effluent from a reforming stage in the
80 combination process, a pressure the same as that prevailing in such reforming stage being maintained during stripping but being reduced before the residual fraction is passed to said fractionating
85 tower.

27. A modification according to Claim 26, wherein the hydrogen and stripped
90 lighter fraction removed from said residual fraction are separated and the hydrogen is recycled to the reforming stage.

28. A modification of the process claimed in any one of claims 1—22, wherein the residual fraction from the
95 crude oil distillation stage instead of being passed as such to said fractionating tower is first subjected to a high temperature thermal treatment at conditions severe enough to cause at least a reduction in viscosity.
100

29. An improved combination process for the refining of petroleum oils substantially as hereinbefore described.

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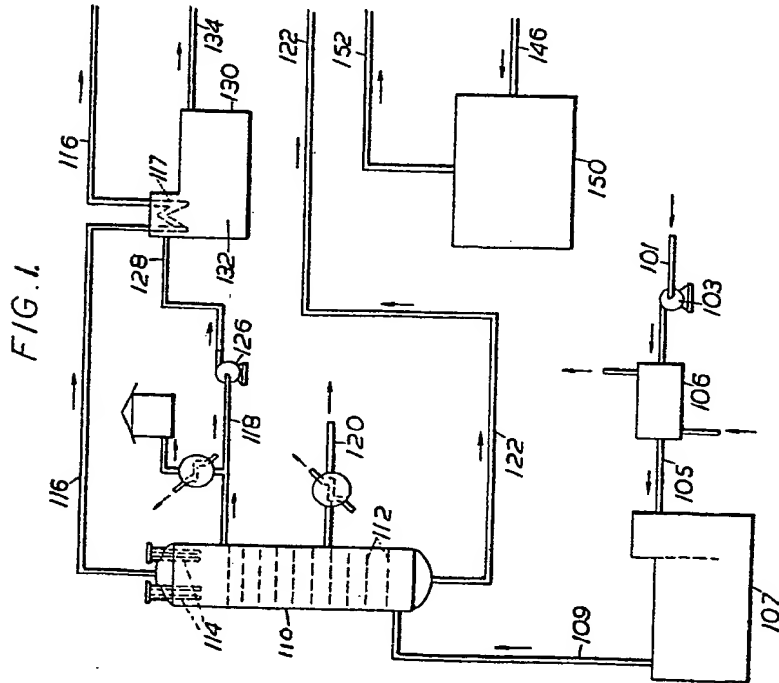
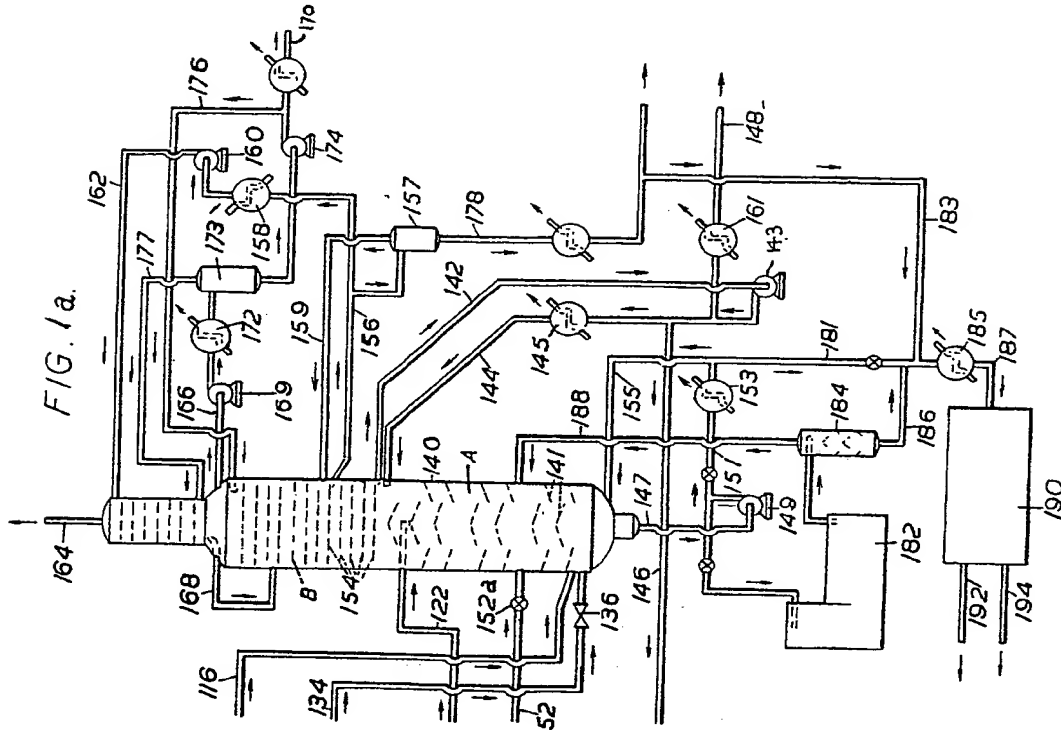
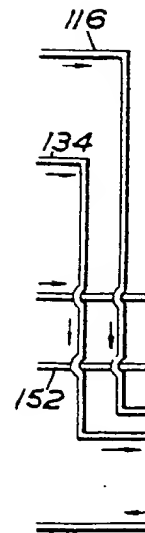
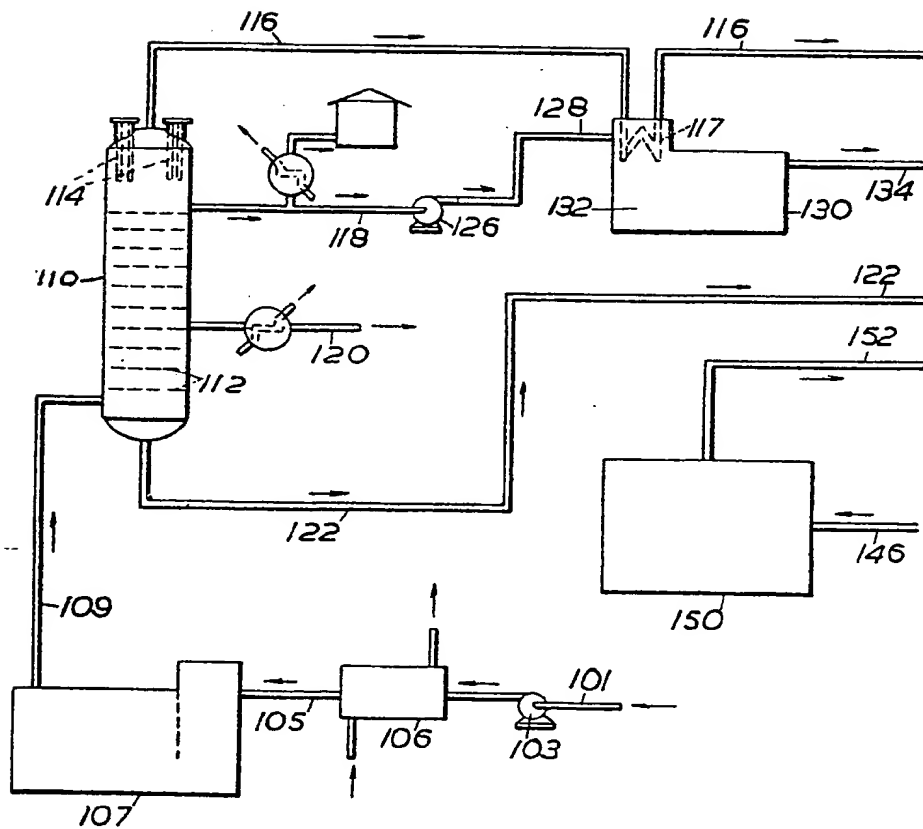


FIG. 1.



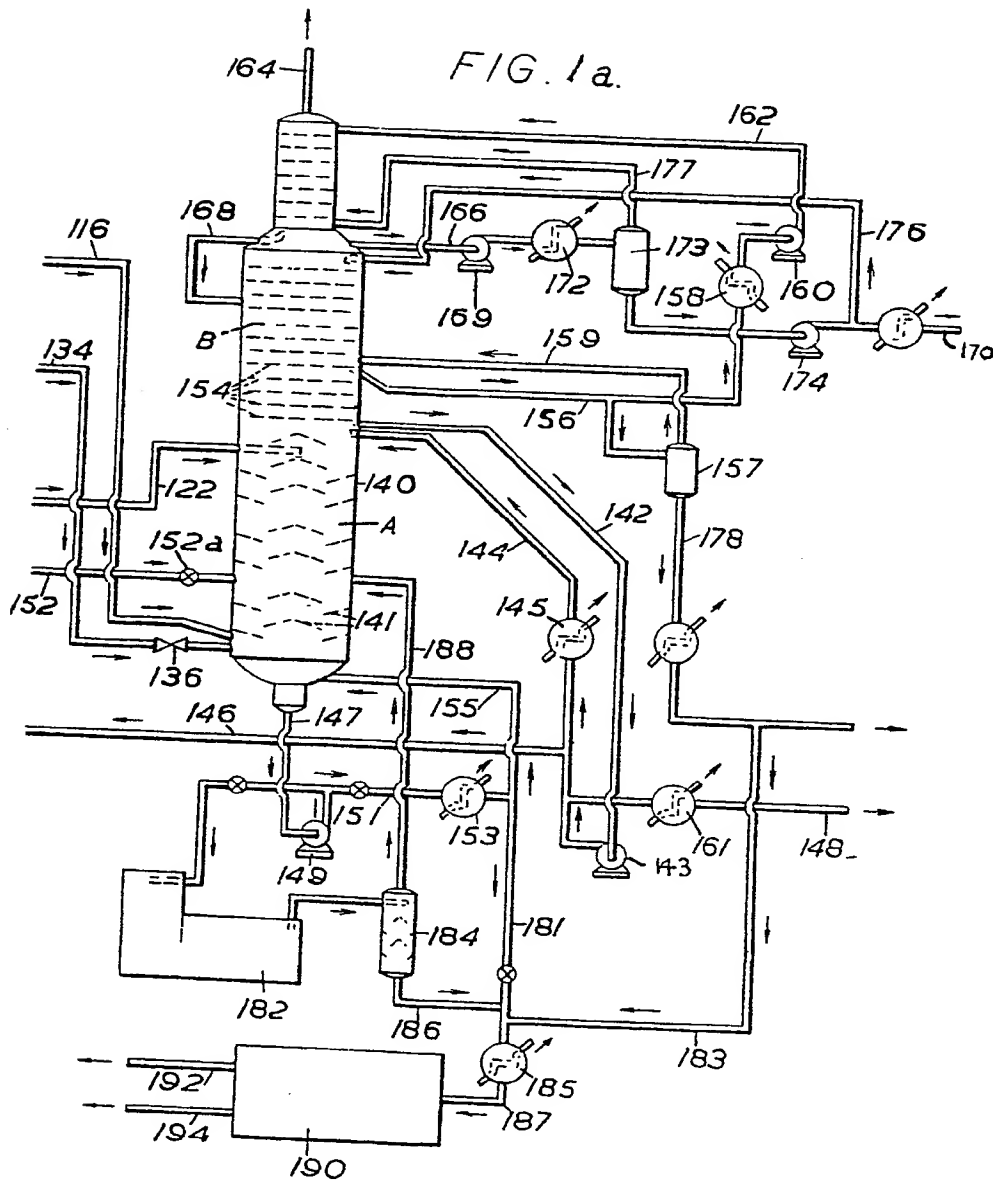


FIG. 2a

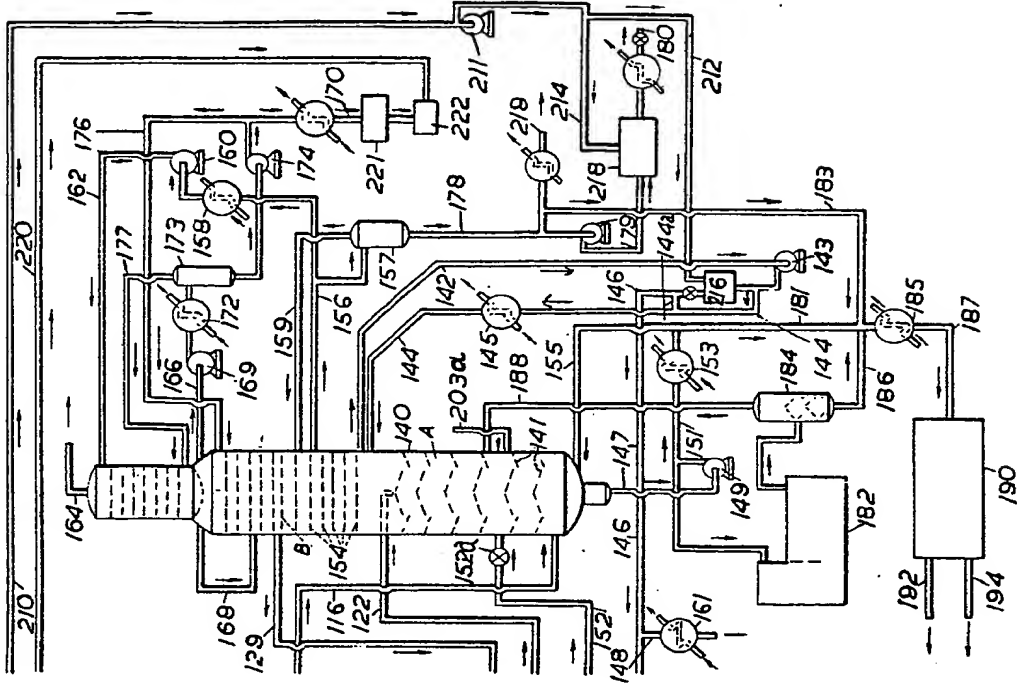


FIG. 2.

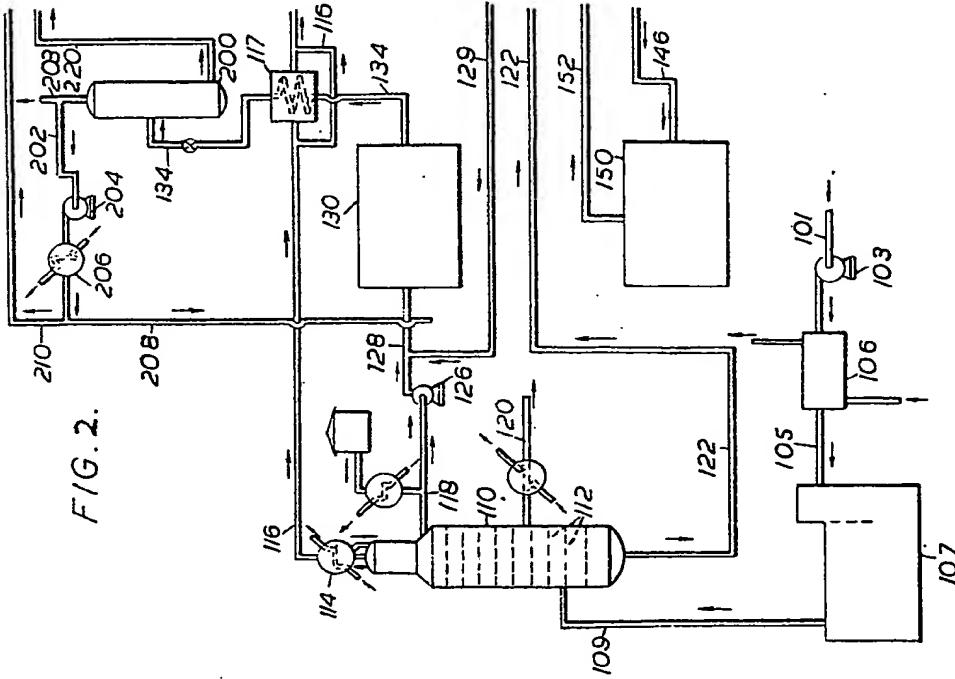
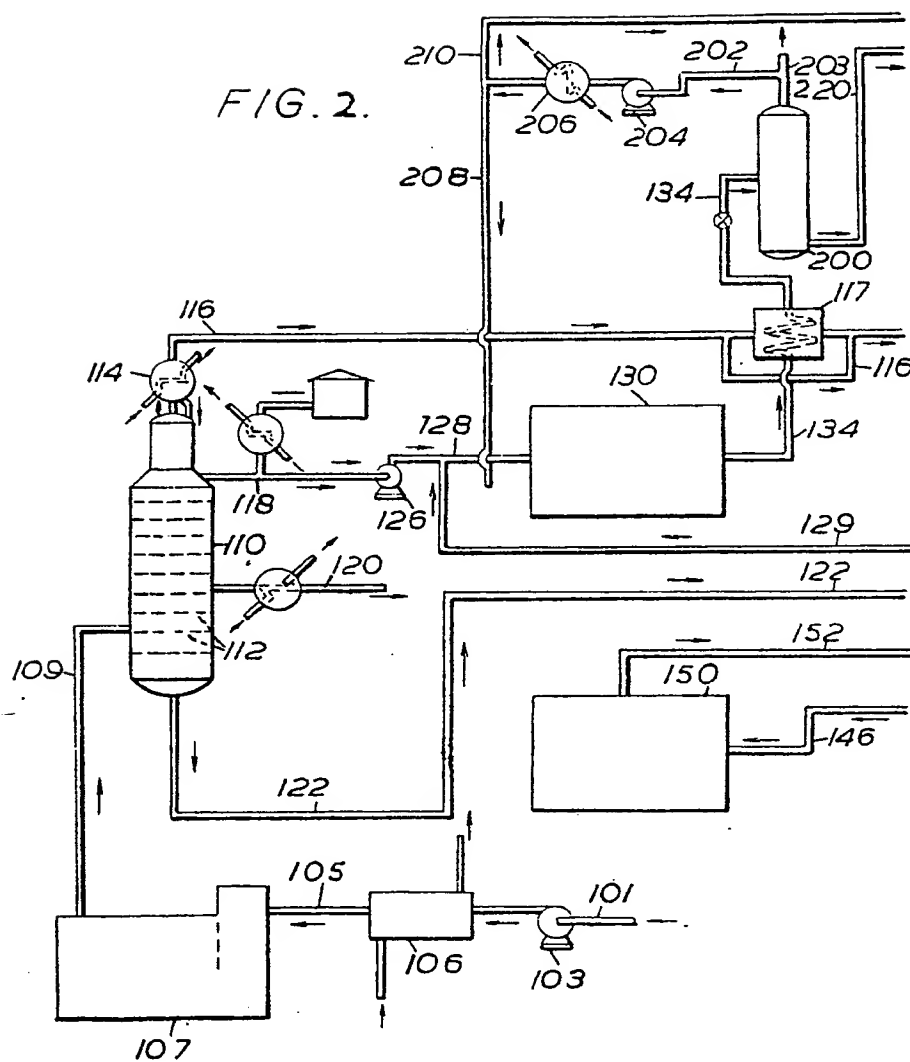


FIG. 2.



2.

16.
129
11
12

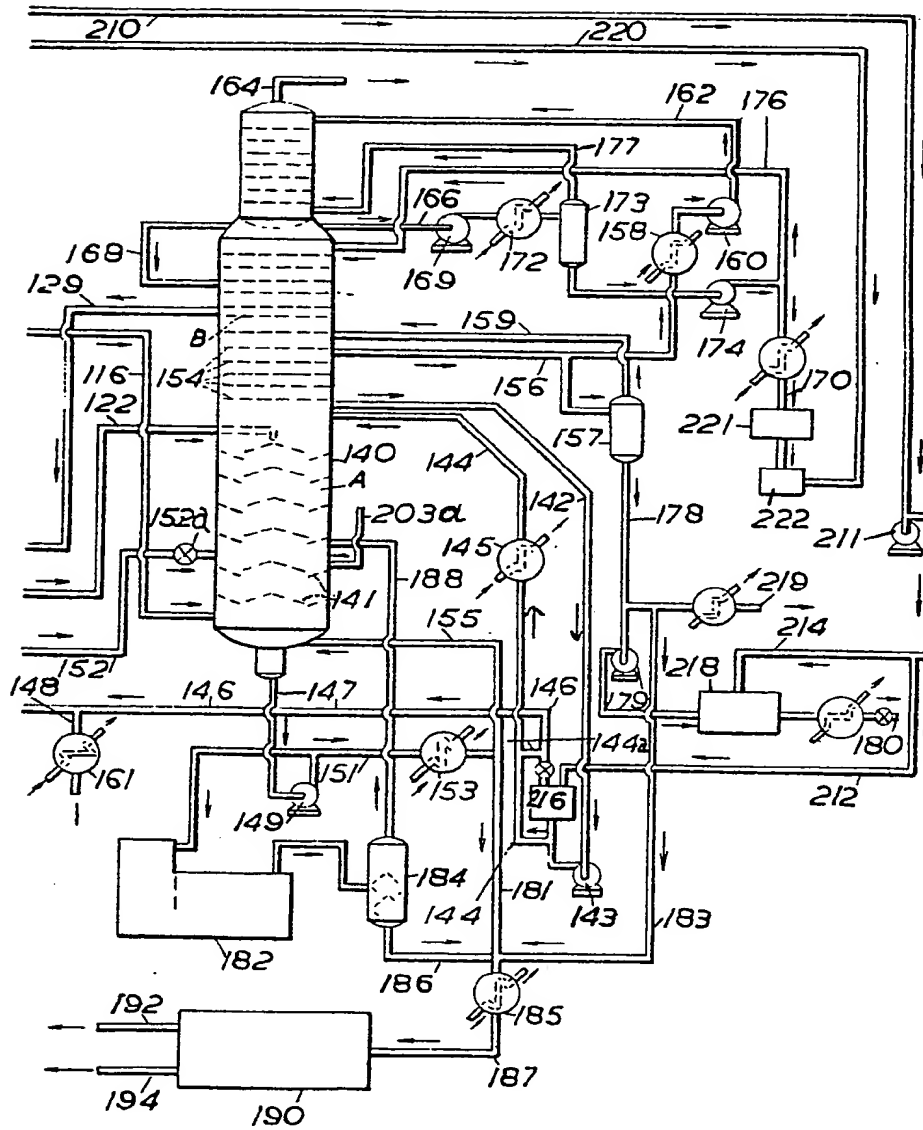
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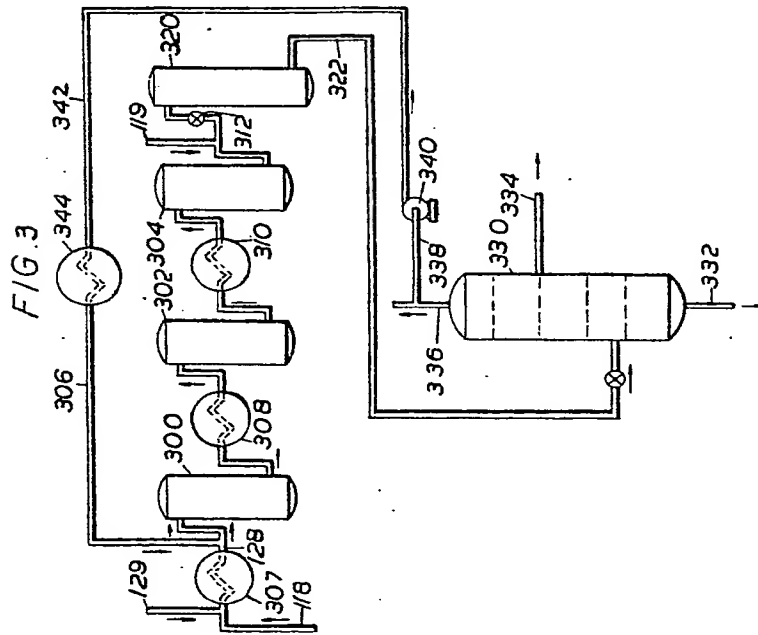
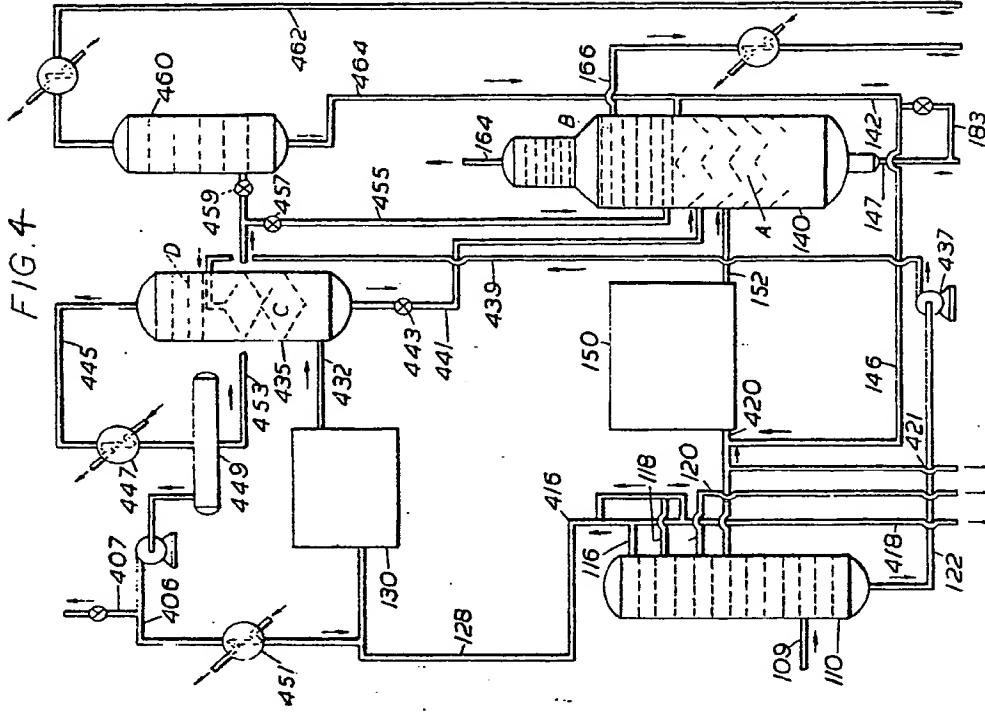
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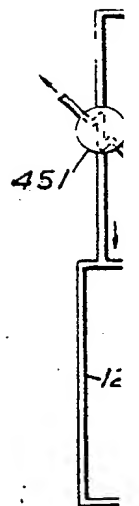
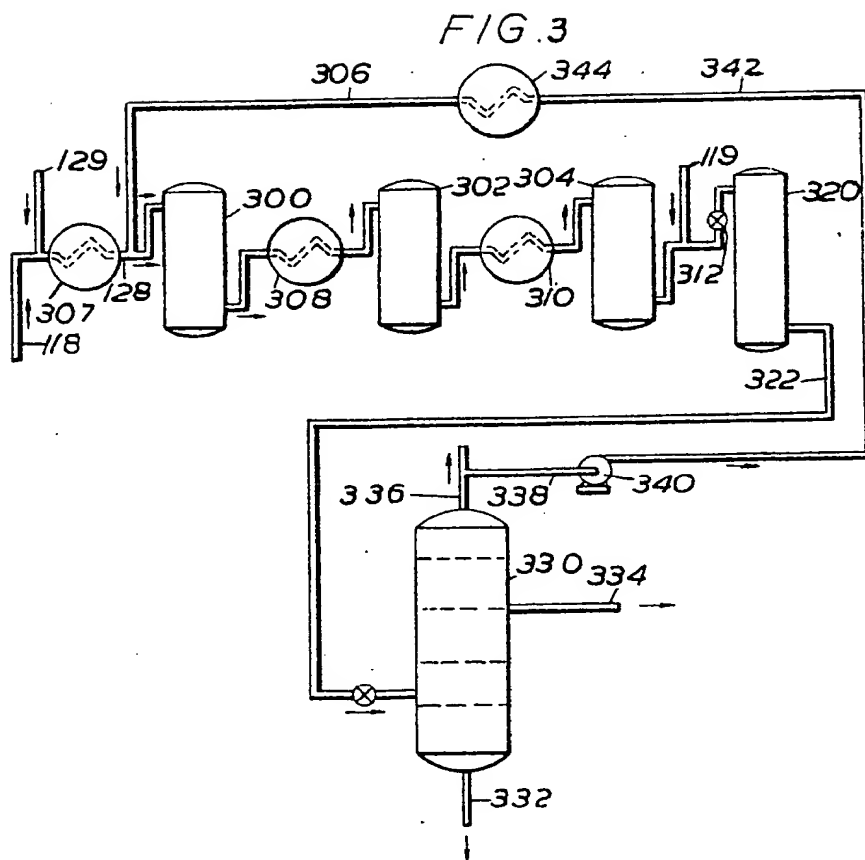
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SHEETS 3 & 4

FIG. 2a







109
110

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SHEETS 5 & 6

FIG. 4

